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Precipitating amino acid solvents for CO_2 capture. Opportunities to reduce costs in Post combustion capture.

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Abstract

Post-combustion CO_2 capture based on absorption and thermal desorption is considered a suitable technology for carbon abatement in fossil fuelled power plants. The economic viability of post-combustion capture has been widely studied. The major drawbacks of this technology are the efficiency penalties and high operating cost related to solvent regeneration and CO_2 compression, and the potential emission of harmful solvent components. Precipitating amino acid solvents are an alternative to conventional amine scrubbing for CO_2 capture from flue gas. These solvents are non-volatile and less toxic than conventional amines. Therefore, the environmental risk related to the emissions of solvent is drastically reduced. Nevertheless, the capture operation with these solvents leads to the formation of precipitates during absorption, which need to be handled in a controlled manner to successfully design a viable process. This work evaluates the technical feasibility of process configurations that contribute to reduce the overall energy use of capture processes based on precipitating amino acid solvents. The process configurations are based on absorption and desorption enhancement effects that are directly or indirectly related to precipitation. The conceptual design of new processes (DECAB and DECAB Plus) based on aqueous solutions of potassium taurate and potassium alanate is described and used to evaluate the performance of these processes. The design is supported by experimental data previously published in the literature. The integration of these process concepts into a coal fired power plant has been evaluated by process simulation of the steam cycle with Aspen Plus ® combined with the capture process requirements estimated by an equilibrium model of the capture process that approximates the thermodynamics of the solvents considered. The DECAB Plus configuration has a net beneficial impact on the energy use of the capture process. The best configuration studied is the DECAB Plus with LVC, which provides up to 4% more power output than the MEA baseline with LVC.

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Keywords:

1. Introduction

Nomenclature

MEA	monoethanolamine
TAU	potassium taurate
ALA	potassium alanate
PCC	post-combustion capture

1.1. Processes based on precipitating amino acid salts

CO_2 absorption in aqueous amino acid salt solutions generally leads to the formation of precipitates, depending on the particular amino acid and the concentration in solution [1, 2]. The composition of the precipitates varies with the structure of the amino acid used to prepare the amino acid salt solution. Some amino acids precipitate the bicarbonate salt enabling the design of desorption systems where the CO_2 is concentrated in the precipitate phase [3]. However, primary amino acids precipitate mainly the pure amino acid [4]. Although CO_2 is not present in the precipitate phase, this phase change can influence the equilibrium reactions in CO_2 absorption and desorption contributing to reduce the energy use of the conventional absorption-desorption capture process. The identified effects related to precipitation are: (1) the enhancement of the specific CO_2 capacity of amino acid salt solutions [5], (2) the acidification of the rich solution, which is not a direct effect of precipitation but is brought about by introducing a phase separation after precipitation that partially removes the supernatant and results in the concentration of the amino acid in the rich solution once the precipitates have been redissolved [6]. The first effect, enhancement of CO_2 absorption, occurs when precipitates are formed during absorption as a result of the chemical reaction of CO_2 with the amino acid salt. The products of this reaction that have a limited solubility (the pure amino acid and / or the HCO_3^-) precipitate when the saturation point is reached. The removal of the solid reaction product from the liquid phase shifts the reaction equilibrium towards the production of more products. The result is a rich stream in the form of slurry that contains mainly the pure amino acid in the solid phase (for primary amino acids) or a mixture of pure amino acid and bicarbonate (for more hindered amino acids) and carbamate, bicarbonate, amino acid salt counter-ion (e.g. potassium, sodium) and remaining amino acid species in the liquid phase [7]. The second effect, acidification of the rich stream, requires the processing of the slurry in a solid-liquid separator to partially separate the supernatant from the solids. This forms a concentrated slurry, which is heated to dissolve the amino acid precipitates. The resulting rich solution, enriched in amino acid, is further processed in the stripper for desorption. The supernatant excess, enriched in amino acid salt counter-ions, is recycled to the absorber without passing through the stripper. After dissolving the amino acid crystals, the pH of the rich stream is decreased before thermal desorption, enhancing the release of the chemically bonded CO_2 [8]. Section 1.3 in this paper describes in more detail the design of a capture process based on precipitating amino acids when the CO_2 is not predominantly present in the precipitate phase.

1.2. Solvent selection

Solvent regeneration in capture processes based on chemical absorption relies on the thermal energy provided by the reboiler and still remains as one of the most important cost factors of the whole capture process. In precipitating

amino acid systems, solvent regeneration also relies on the thermal energy provided by the reboiler but also on the pH-shift effect described in the previous section, which helps to the reduction of this energy. Both thermal and pH-shift effects in these processes are influenced by specific amino acid properties. There are key solvent characteristics that are of importance for its application in a thermal absorption – desorption process, such as the net CO_2 capacity of the solvent, the absorption rates, the heat of absorption, and the heat capacity of the solvent [9]. In addition to these, the extension of the precipitation reaction during absorption is important to the reduction of the regeneration energy, since the formation of more amino acid precipitates helps to decrease the pH of the new rich stream formed when a fraction of the supernatant is recycled. The precipitates need to be dissolved for the pH-shift effect to be effective. Therefore, the amino acid solubility and heat of dissolution are also important properties for the selection of a solvent for these processes. Table 1 shows a summary of the amino acids considered most relevant for this application based on these properties.

Table 1. Amino acid properties extracted from references [7-11] for precipitate type and vapour liquid equilibrium data, from [12,13] for melting point and solubility.

Amino Acid Concentration [M]	Formula	Abbreviation	Mw [g/mol]	Melting point [ºC]	Solubility [g/100mL]	Precipitate type
Taurine 4M	C2H7NO3S	TAU	125	328.0	9.49	Amino acid
α-alanine 4M	C3H7NO2	ALA	89	316.5	16.58	Amino acid and bicarbonate
2-aminoisobutyric acid (2-methylalanine) 4M	C4H9NO2	AIB	103	335.0	12.06-18.40	Amino acid
6-aminohexanoic acid (ε-aminocaproic acid) 6M	C6H13NO2	6-AHA	131	205.0	50.50	Bicarbonate

Based on the measurement of CO_2 absorption equilibrium isotherms (40 and 120ºC) and of solutions of the amino acids listed in Table 1 within the precipitation regime, Sanchez Fernandez *et al* show an increasing absorption capacity in the order 6-AHA > ALA > AIB > TAU [8]. With respect to regeneration, the 6-AHA and AIB solutions show lower partial pressures of CO_2 than taurine and alanine at 120ºC, which implies that more stripping steam will be required to regenerate these solvents with the consequent impact on the thermal regeneration energy. The critical points for solvent precipitation were also determined by Sanchez Fernandez *et al* [8] and they decrease from 0.6 mol CO_2 / mol AmA to 0.15 mol CO_2 / mol AmA in the order 6-AHA > TAU > ALA > AIB. The critical point for precipitation has an effect on the application of these solutions to the precipitating processes. The formation of precipitates at high CO_2 partial pressure (and loading) decreases the enhancement effect during absorption. On the other hand, the formation of precipitates at very low CO_2 partial pressures (and loadings) will required a deep regeneration to keep the lean loading below the critical point for precipitation. Considering the absorption, regeneration and precipitation characteristics, the most promising amino acid in Table 1 are taurine and α-alanine. Therefore, these two solvents were selected for the feasibility study of the precipitating amino acid capture process.

1.3. Process concept description (DECAB and DECAB Plus with LVC)

Figure 1 illustrates a generic precipitating amino acid process scheme based on the potassium salt of taurine that exploits the effects described in section 1.1. The flue gas, containing circa 13 %v/v CO_2 at 40ºC, is treated with a solution of potassium taurate with 1:1 molar amounts of taurine and KOH in a suitable contactor that can handle solids. This can be a spray tower contactor or a sequence of spray tower contactor and packed column. The slurry that results from the absorption of CO_2 can be directly regenerated in the stripper, via the lean-rich heat exchanger, in a process concept known as DECAB. In this case, the separator in Figure 1 is not needed. Alternatively, the slurry

can be treated in the solid-liquid separator yielding two streams; new rich stream and recycle. The new rich stream is a slurry enriched in taurine, that has a taurine-to-potassium molar ratio ([TAU/K]) higher than 1. This stream is further processed in the stripper, where the CO_2 is released, yielding the new lean stream. By re-dissolving the amino acid crystals, the pH of the new rich solution is decreased before thermal desorption, which results in higher partial pressures of CO_2 at stripper conditions that facilitate desorption. The recycle stream is the supernatant excess enriched in potassium ions, that has a taurine-to-potassium molar ratio lower than 1. This stream is mixed with the new lean stream and is fed back to the absorber without passing through the stripper. Mixing the recycle and the new lean stream results in the initial lean stream that has a taurine-to-potassium molar ratio of 1. This process alternative is known as DECAB Plus.

The lean vapour compression (LVC) concept is an option to reduce the regeneration energy that has been described and investigated for other solvent systems, such as MEA [14]. In this process concept, illustrated in Figure 2, the lean stream that leaves the stripper column is flashed adiabatically in a separate vessel. As a consequence of the pressure reduction, the solvent is partially evaporated. The vapour generated is re-compressed and re-injected at the base of the stripper column. The benefits of this process concept are that part of the CO_2 remaining in the lean stream is flashed, reducing the lean loading and that more steam is generated, reducing the necessary heat input to the reboiler.

2. Conceptual design and modelling

2.1. Capture process modelling

An equilibrium process model (VLEMS) has been developed for aqueous solutions of amino acids and MEA and has been used for the evaluation of the different process configurations proposed in this work [7, 8]. The absorption and desorption performances are estimated by a series of equilibrium stages where the mass balances, enthalpy balance and equilibrium relations are simultaneously solved providing the temperature and composition profiles and the reboiler energy. The model is based on the following assumptions:

- The only components that are considered in the VLE calculations are CO_2 and water. Amino acids are non-volatile and in the case of amines, the vaporization is neglected.
- Each stage is assumed to be well mixed and process conditions and gas and liquid composition are uniform in each stage.
- Temperature, pressure and material equilibrium is reached in each stage.
- The reboiler is considered as an equilibrium stage and heat losses are not considered.

The Vapor-Liquid-Solid Equilibrium (VL(S)E) data for the amino acid salts under investigation are necessary to estimate solid-liquid partition and Vapor-Liquid Equilibrium (VLE) at different conditions. In previous work [7], a conceptual design methodology was introduced, which required three different experimental tests: the pH determination of solutions containing different concentrations of amino acid and potassium hydroxide, the measurement of the VLE on those solutions at different temperatures and the determination of the weight of solids and solid type formed during CO_2 absorption of those solutions. The results of each experiments were fitted to empirical expressions that provided the estimation of the VL(S)E at different conditions [7,8].

In the equilibrium model, a constant value of 0.8 Murphree efficiency is applied at every stage for CO_2. This value was selected to fit the experimental results of Knudsen [15] for MEA, which were conducted at the pilot plant of Dong at Esbjerg power plant station. For the other solvents (potassium taurate and alanate), the same value of the Murphree efficiency was used due to the lack of pilot plant experimental results with these solvents. This assumption is equivalent to assuming similar absorption rates to MEA for these two solvents and could lead to loadings that deviate from those attainable at real operating conditions. Although this could be a limiting factor for future process implementation, it was considered acceptable for the conceptual design because the objective is to screen multiple solvents and process configurations. The model assumptions could be checked in future studies for an optimal solvent and process configuration. For a detailed description of modelling assumption the reader is referred to previous studies [7, 8].

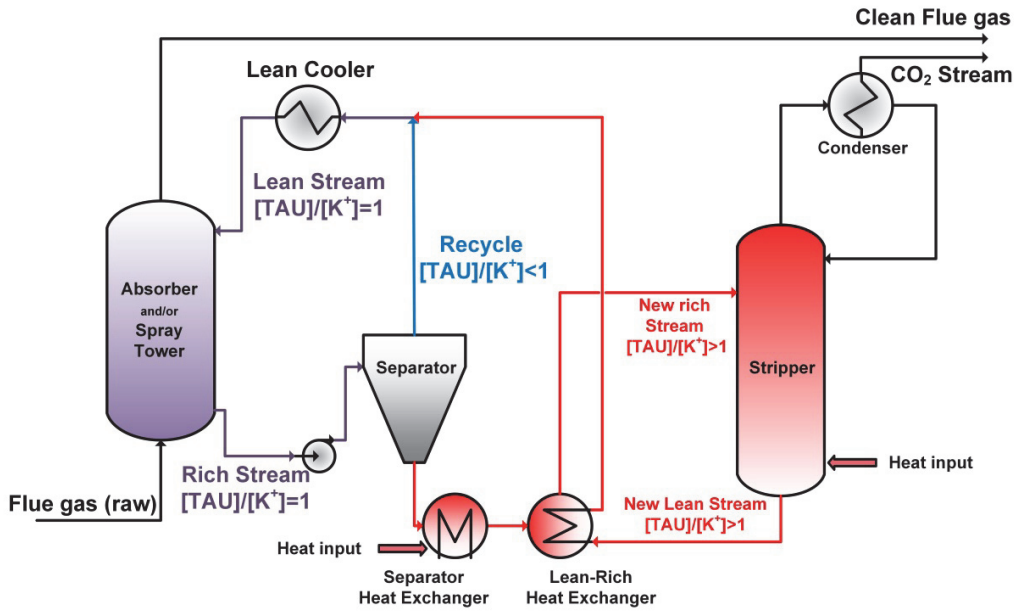

Fig. 1. Conventional precipitating amino acid based process for CO_2 capture. Besides the heat input necessary to regenerate the solvent, in the case of precipitating amino acids two more effects are possible: Enhanced absorption (takes place in the purple region) due to the precipitation of reaction products during absorption and enhanced desorption (takes place in the red region) due to a lower pH that results from increasing the amino acid to of K^+ ratio in solution.

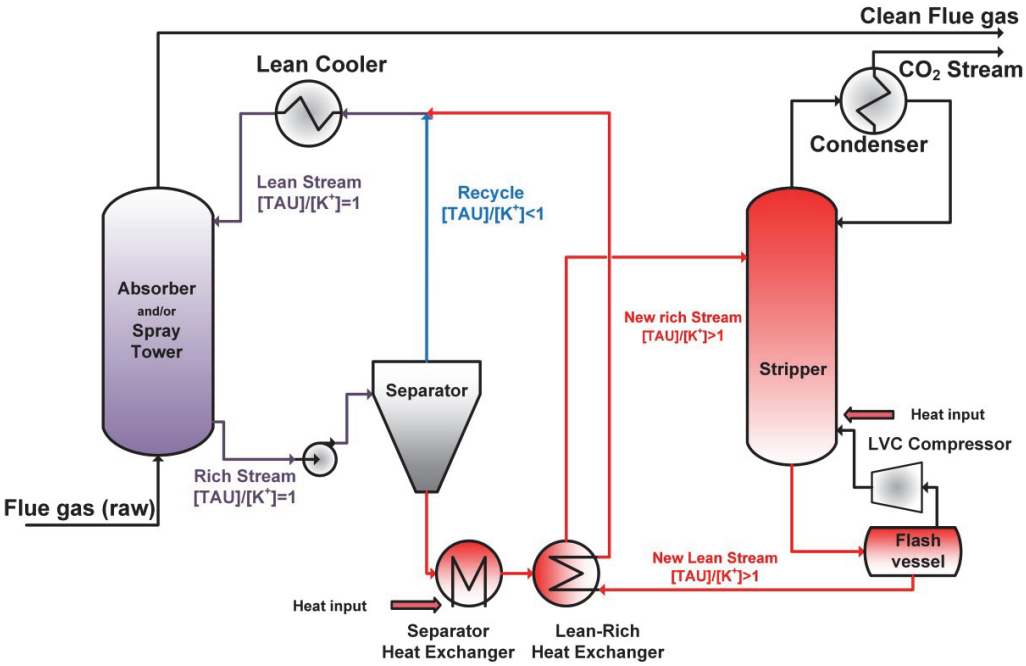

Fig. 2. DECAB Plus process based on precipitating potassium taurate with LVC. The ratio of the total taurine concentration to the total potassium concentration in solution is indicated below the name of each stream. This ratio can be 1 (equimolar solution), higher than 1 (solution with less potassium than the equimolar) or less than 1 (solution with more potassium than the equimolar).

A number of process variables are important to the evaluation of the DECAB and DECAB Plus process concepts:

- Taurine-to-potassium molar ratio ([TAU/K]) represents the ratio of the total concentration of taurine to the total concentration of potassium, which is equal to the concentration of KOH used to prepared the amino acid salt solution.
- Initial pH of taurine – KOH solutions (pH_0): represents the pH of an aqueous solution of taurine and KOH measured at 40ºC. It is related to the [TAU/K] as shown in our previous work:

  $$\mathrm{pH}_0 = f\left(\frac{[TAU]}{[K^+]}\right) \quad (1)$$

- Recycle split fraction (RSF): represents the fraction of the liquid supernatant (excluding the solid phase) that is recycled to the absorber column.

  $$RSF = \frac{L_R}{L-S} \quad (2)$$

  Here, L_R represents the flow (kmol/s) of the recycle stream, L represents the flow (kmol/s) of the total rich stream leaving the absorber and S represents the flow (kmol/s) of the solid phase.

2.2. Integration into the steam cycle of a coal fired power plant

The integration of the process concepts into the steam cycle of a coal fired plant has been evaluated by process simulation of the power plant with Aspen Plus ® combined with the capture process requirements estimated by the equilibrium model that approximates the thermodynamics of the solvents considered. The power plant is based on an Advanced Super-Critical (ASC) Boiler and Turbine with a gross power output of about 625 MWe without carbon capture. The flue gas that exits the boiler goes through state-of-the-art air pollution control systems before being fed to the DECAB Plus capture process. The steam cycle, depicted in Figure 3, is a triple pressure cycle with re-heat. The live steam parameters are 298 bar and 600ºC and the re-heated steam conditions are 64.5 bar and 620ºC. The steam turbine plant consists of HP turbine, IP turbine and LP turbine with extraction points for regenerative heating of feed water and condensate. There are seven feed water heaters that produce boiler feed-water at 309.6ºC.

The steam extraction system needed to provide the DECAB capture plant with process steam for solvent regeneration is inserted into the IP/LP crossover. The investigated cases, assumed a new build power station with carbon capture; it is therefore considered that they would be designed simultaneously and the steam turbines would be optimized for steam extraction at nominal conditions. At the IP/LP crossover the nominal design pressure with CO_2 capture is 4 bar. As the power plant is considered a base load power plant, the effect of the turbine re-design and steam extraction on the efficiency at part load has not been considered.

Moreover, in the DECAB Plus process the slurry feed to the stripper requires the input of low-grade energy to dissolve the amino acid crystals. The DECAB Plus process, depicted in Figure 1, uses the energy from the hot lean stream to heat the rich stream that is fed to the stripper. The additional energy input required for the phase change is approximated by the dissolution energy, which is supplied in the separator heat exchanger. For this purpose various options could be considered:

- Using saturated steam at the same pressure as the reboiler (Figure 3): this option has been proven to be highly inefficient since the temperature level needed in the separator heat exchanger is below 80ºC. Therefore, the use of saturated steam results in a substantial penalty to the power output of the steam cycle
- Using hot condensate return from the reboiler with additional hot water from the feed water train (Figure 4): This option is more efficient than the previous one because it does not require dragging more sulperheated steam from the IP/LPcrossover.

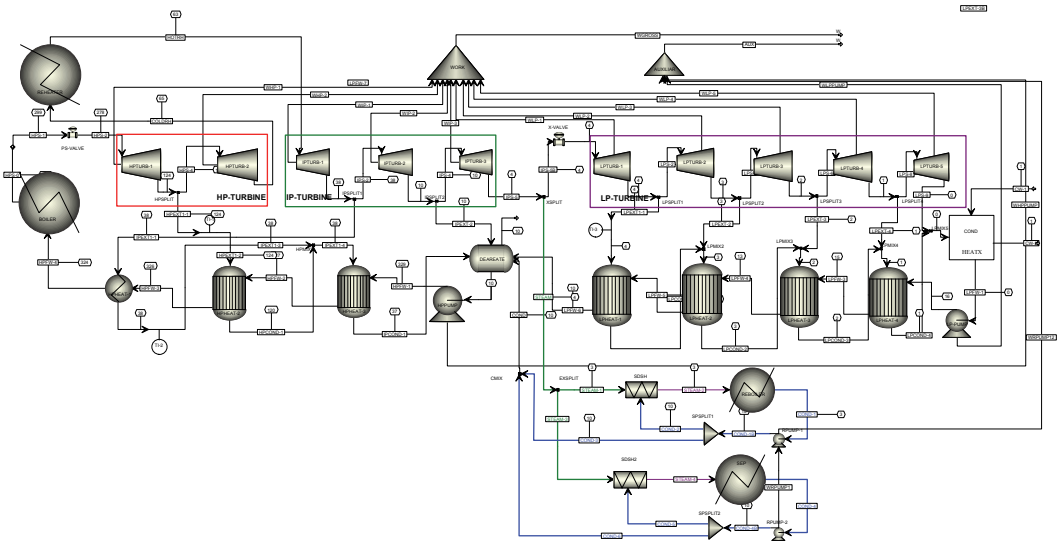

Fig. 3. Integration of the DECAB process concept for CO_2 capture into a supercritical steam cycle with the use of condensing steam for the reboiler and hot returned condensate for the separator

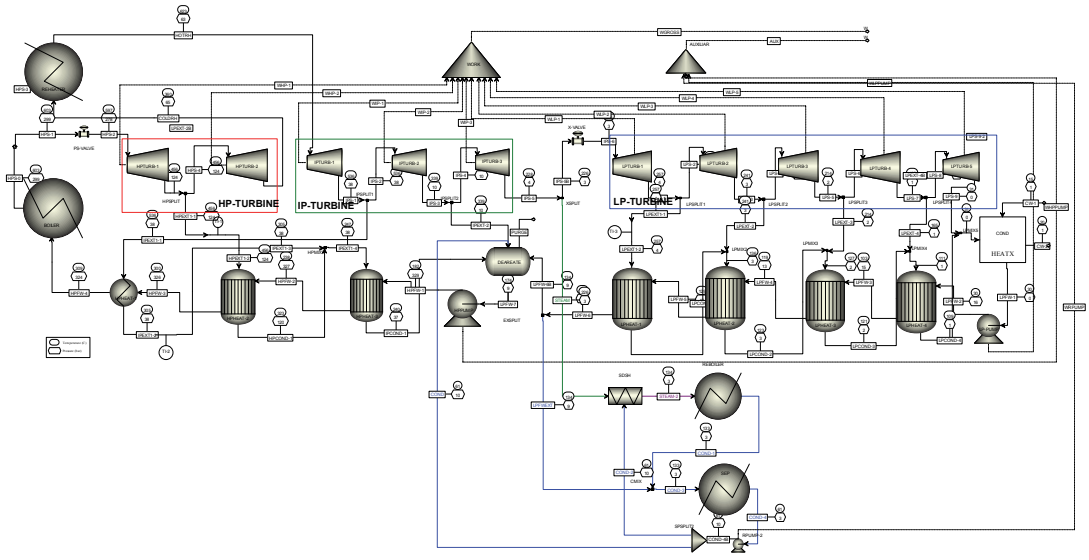

Fig. 4. Integration of the DECAB process concept for CO_2 capture into a supercritical steam cycle with the use of condensing steam for the reboiler and hot returned condensate for the separator

The steam extracted from the crossover needs to be conditioned to generate saturated steam at the required reboiler pressure. The reboiler requires a steam pressure of 3 bar. The steam is then suitably conditioned (through pressure reduction and direct spray injection attemperation) for reboiler use. Steam conditions at the reboiler entrance are 134ºC and 3.05 bar. The condensate is returned into the boiler feed water train or used for dissolving taurine crystals in the separator heat exchanger.

3. Results and discussion

3.1. DECAB Plus process optimisation

The optimization of the DECAB Plus capture process has been done in a stepwise approach. In a first stage, the lean loading without any recycle (DECAB configuration) was optimised to minimize reboiler duty. This optimization is shown in Figure 5 at different reboiler temperatures for the TAU and ALA solvents. The minimum specific reboiler duty occurs at a CO_2 loading of 0.26-0.27 mol CO_2/mol TAU at any temperature for the TAU solvent (Figure 5 left). In the case of the ALA solvent (Figure 5 right), the minimum specific reboiler duty occurs at a CO_2 loading of 0.32-0.33 mol CO_2/mol ALA. At this loading the specific reboiler duty for potassium alanate is lower than for potassium taurate (also included in the figure). Nevertheless, for loadings above 0.26 mol CO_2 / mol ALA, precipitates will form at temperatures around 40ºC. Therefore, the loading needs to be restricted at that value to avoid precipitation of the lean stream when is fed to the absorber. The minimum reboiler duty is then 3.25 GJ/t CO_2, with a reboiler temperature of 130ºC. The effect of the reboiler operating temperature is not very substantial but is appreciable from the results in Figure 5. Higher reboiler temperatures lead to lower specific reboiler duties due to a more favorable water to CO_2 ratio at stripper's top stage conditions.

The second step in the optimization of the DECAB Plus process is the optimization of the recycle split fraction (RSF). This step was only performed with the TAU solvent due to lack of data for the ALA solvent. For every temperature analyzed in Figure 5, the reboiler pressure was fixed at the value that provides the optimum lean loading. Subsequently, the recycle split fraction (RSF) was varied from 0% to 50% in order to induce the pH-shift effect of the DECAB Plus process configuration. Figure 6 (left) shows the optimization of the RSF at different reboiler temperatures. The DECAB Plus process configuration reduces the specific reboiler energy by 27-29% depending on the operating reboiler temperature. For all temperatures, the value of the RSF that minimizes the specific reboiler duty is around 30%. For higher values of the RSF, the net capacity of the solvent is considerably reduced due to the recycle of the absorbed CO_2 back to the absorber column. For these cases, the solvent flow needs to be increased substantially, to maintain the CO_2 removal percentage at 90%, leading to an increase in the required reboiler duty. For the DECAB Plus configuration, lower reboiler temperatures lead to lower specific reboiler duties, despite the fact, that for the DECAB configuration (i.e. 0% RSF) the opposite behavior was found. This change can be explained by the variation of the pH-shift effect with temperature. Figure 6 (Right) shows the influence of pH_0 in the equilibrium partial pressure of CO_2 at optimum lean loading conditions. When the pH_0 is decreased, the equilibrium partial pressure of CO_2 increases significantly. However, for higher values of temperatures the effect becomes less prominent, as indicated by the reduction in the slope of the PCO_2 vs. pH_0 lines represented in Figure 6. Therefore, the RSF has a more markedly effect on the reboiler duty at lower temperatures than at higher temperatures.

For the evaluation of the LVC addition to the DECAB and DECAB Plus concepts, several simulations were performed at different reboiler temperatures. The reboiler pressure was kept constant at the value that minimized the specific reboiler duty for every temperature. The pressure in the flash vessel was decreased gradually, starting from a value equal to the reboiler pressure, to atmospheric conditions. The results of these simulations are shown in Figure 7 (Left) for reboiler temperatures of 120ºC and 130ºC. Figure 7 shows the specific energy requirements of the reboiler, which include the energy required to power the compressor in the form of steam that would result in a power loss to the turbine equivalent to the LVC compressor power. When the rich stream is flashed, the reboiler duty necessary to regenerate the solvent is reduced due to the extra steam generated. The extra steam is then compressed to the reboiler pressure and fed back to the stripper. As shown in Figure 7, the total energy required for regeneration decreases with decreasing flash pressures. However, for flash pressures near atmospheric conditions, the energy required by the compressor becomes a dominant contributor to the total energy requirements and, therefore, the specific energy requirements raise. The implications of the use of compressor work to reduce the energy requirements are evaluated rigorously by simulation and are discussed in section 3.

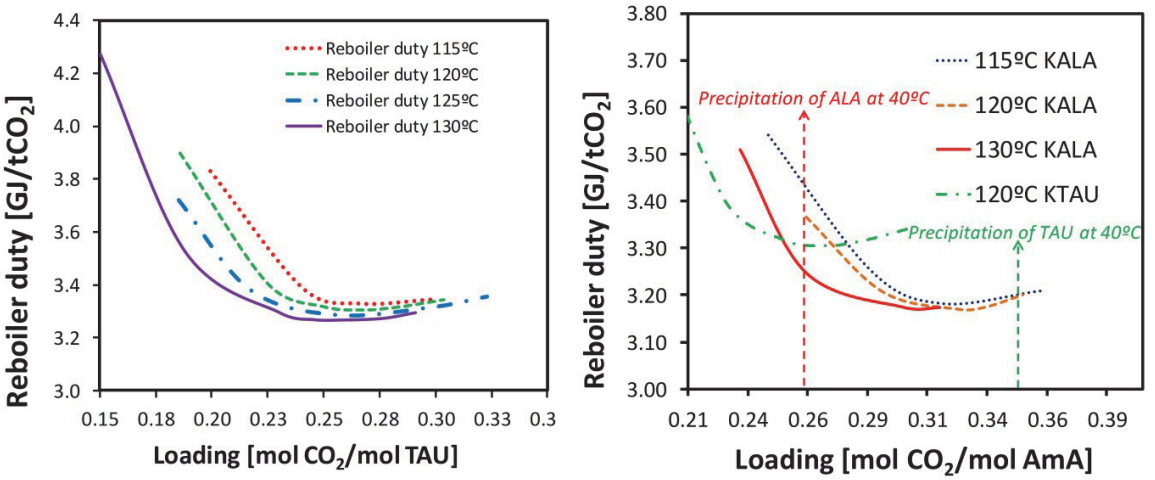

Fig. 5. (Left) Optimisation of specific reboiler duty with respect to reboiler temperature for potassium taurate and DECAB process configuration illustrated in Figure 1. Simulations were performed at a constant recycle split fraction (RSF) of 0% and constant CO_2 removal of 90%. (Right) Optimisation of specific reboiler duty with respect to reboiler temperature for potassium alanate and DECAB process configuration illustrated in Figure 1. Simulations were performed at a constant recycle split fraction (RSF) of 0% and constant CO_2 removal of 90%.

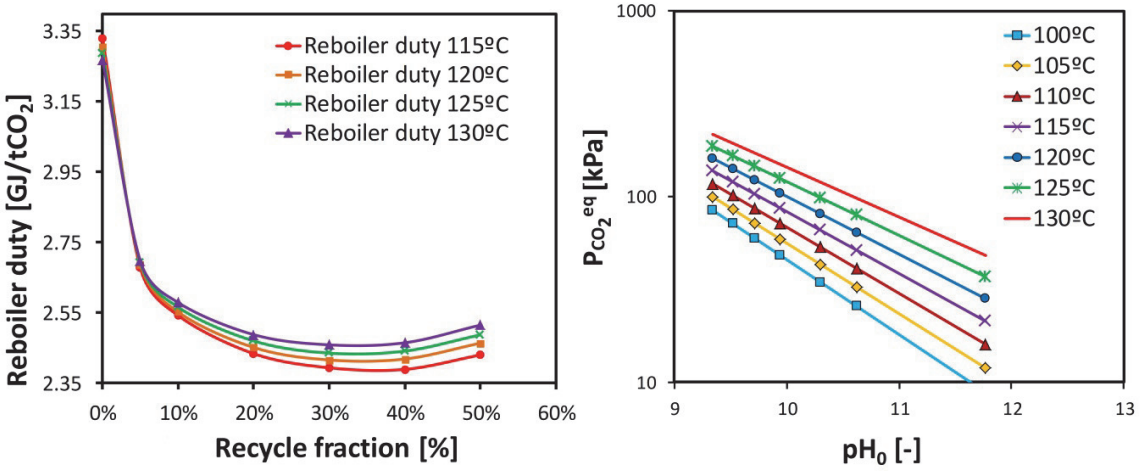

Fig. 6. (Left) Optimisation of the reboiler duty with respect the recycle split fraction (RSF) at different temperatures for the DECAB Plus process configuration illustrated in Figure 1 and the potassium taurate solvent. Simulations were performed at the optimum lean loading for every temperature and a constant CO_2 removal of 90%. (Right) Influence of pH_0 in the equilibrium partial pressure of CO_2. Simulations were performed at a constant loading of 0.26 mol CO_2/mol TAU and different temperatures.

The influence of LVC on the DECAB Plus process configuration was also investigated at constant reboiler temperatures and pressures. Figure 7 (Right) shows the influence of the LVC flash pressure on the specific regeneration energy at 120ºC and different values of the RSF. The effect of flashing the lean stream on the total energy required to regenerate the solvent is similar for every value of the RSF. In relative terms, the value of the initial specific energy is reduced from 7.9% to 10.0% when RSF increases from 0% to 40%.

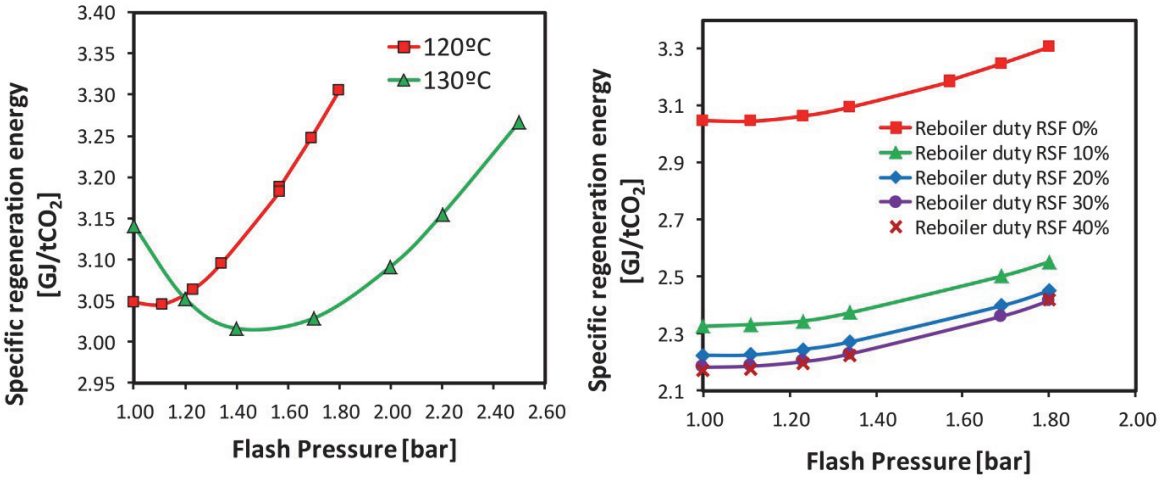

Fig. 7. (Left) Optimisation of LVC flash pressure for the DECAB process concept with LVC illustrated in Figure 2. Simulations were performed at constant reboiler pressure corresponding to the optimal value for every temperature, no recycle split fraction (0% RSF) and a constant CO_2 removal of 90%. (Right) Effect of LVC flash pressure on the specific regeneration energy for the DECAB Plus process concept with LVC illustrated in Figure 2. Simulations were performed at constant reboiler temperature (120ºC) and a constant CO_2 removal of 90%.

4. Process integration and evaluation

This section provides an overview of the process concepts analysed in this work and the integrated performance of a supercritical steam cycle with the DECAB and the DECAB Plus processes. Table 2 shows the performance of the process configurations investigated and the most optimal process conditions of each configuration. The evaluation of a MEA capture process, based on the same model and assumptions as in the other process configurations, has been added to the table for comparison [7, 8]. The evaluation is based on the net power output of the steam cycle of each configuration, which has been calculated rigorously with the model described in section 2.2 and includes the energy use of each process configuration, the work of the LVC compressor (when present), the work to compress the CO_2 to 110 bar and the auxiliary work for pumps.

The optimisation of operating conditions was based on the reduction of reboiler energy. Based on potassium taurate, the DECAB process configuration reduces the reboiler duty but at the expense of increasing the separator energy. Under the integration conditions investigated there is a marginal improvement in the overall energy use of the DECAB configuration as compared to conventional MEA when the low pressure feed water is used in conjunction with the returned condensate to provide the energy required by the separator. This leads to an increase in the power output of the steam cycle of about 4 MW. The reduction in the overall energy required by the DECAB process when the solvent is changed from taurine to alanine is not substantial. The major savings with respect to taurine rely on the lower heat of dissolution of alanine, resulting in lower energy required by the separator to dissolve the precipitates.

The addition of the LVC to the process reduces the overall energy requirements of the DECAB process configuration. In the case of potassium alanate, the required energy is lower than the MEA baseline with an extra 2.4% in power output over the power obtained by adding the LVC to the MEA conventional process configuration. Nevertheless, the DECAB Plus configuration has a net beneficial impact on the energy use of the capture process. The best configuration is the DECAB Plus with LVC, which provides up to 4% more power output than the MEA with LVC.

Table 2. Comparison between the investigated process concepts, based on 4M potassium taurate, 4M potassium alanate and conventional MEA.

Process configuration	Solvent	Reboiler			S-L separator[b]		LVC	Energy		Work		Auxiliary	Power output	
		T	*P*	*Ps*	*T*	*RSF*	*P*	Q_R	Q_{Sep}	W_{LVC}	$W_{Comp.}$	*overall*	*Gross*	*Net*
		ºC	bar	bar	ºC	%	bar	GJ/t	GJ/t	kWh/t	kWh/t	MW	MW	MW
No Capture	NA											24.4	625	**600.6**
Conventional	MEA[a]	120	1.9	3.13	NA	NA	1.9	3.66	0.00	0.0	93.6	56.8	539.9	**483.0**
DECAB	TAU	120	1.8	3.13	40	0%	1.8	3.31	1.01	0.0	95.2	57.6	544.3	**486.8**
DECAB	ALA	120	1.8	3.13	40	0%	1.8	3.37	0.56	0.0	95.2	57.4	544.9	**487.5**
Conventional + LVC	MEA	120	1.9	3.13	NA	NA	1.2	3.18	0.00	9.2	93.6	60.1	549.2	**489.2**
DECAB + LVC	TAU	120	1.8	3.13	40	0%	1.1	2.86	0.93	12.1	95.2	61.8	553.6	**491.8**
DECAB + LVC	ALA	130	2.6	4.16	40	0%	1.4	2.52	0.54	18.2	84.2	60.0	561.0	**501.0**
DECAB Plus	TAU	120	1.8	3.13	40	20%	1.8	2.45	0.96	0.0	95.2	57.7	562.1	**504.4**
DECAB Plus + LVC	TAU	120	1.8	3.13	40	20%	1.1	2.07	0.88	9.9	95.2	61.2	570.0	**508.8**

5. Conclusions and future work

DECAB and DECAB Plus are two novel process concepts based on precipitating amino acids that can reduce the regeneration energy required by the CO_2 capture process. An analysis of process conditions and different configurations of these two process concepts has been conducted based on the necessary energy to regenerate the solvent. The configurations analysed contribute to decrease this energy. However, the developed process configurations also require low-grade energy to dissolve the precipitates formed during absorption. An integration scheme with a supercritical steam cycle has been proposed that uses the energy of the reboiler condensate and part of the energy of the boiler feed water train to provide the necessary energy input to re-dissolve the crystals.

All the configurations analysed reduce the overall energy use of the capture process with respect to the MEA baseline. The best process configuration in this study is the DECAB Plus process configuration based on potassium taurate with a reboiler temperature of 120ºC and solid-liquid separation at 40ºC. This alternative reduces the overall energy of the capture process by 15% compared to the MEA baseline with compression of the CO_2 to 110 bar. This results in an extra 4% power output with respect to the MEA baseline.

Other amino acid solvents were investigated in this work in order to explore potential reductions in the energy required by the DECAB process configuration. Among the solvents investigated, the potassium salt of α-alanine was identified as a potential candidate for the application in the DECAB configuration. The performance of the DECAB configuration based on potassium alanate is better than potassium taurate due to the lower heat of dissolution of α-alanine. With the addition of LVC it increases the power output of the baseline with LVC by 2.4%. Considering the benefits that the DECAB configuration with potassium alanate has over potassium taurate, a DECAB Plus configuration based on this solvent has the potential to further reduce the energy requirements of the capture process. Moreover, a better integration of the returned condensate within the steam cycle should reduce the energy figures presented in this work.

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